

**EUROPEAN PATENT SPECIFICATION**

(12)

(45) Date of publication of patent specification: 15.11.89

(21) Application number: 83102890.7

(22) Date of filing: 23.03.83

(30) Divisional application 87106294 filed on 30.04.87.

(51) Int. Cl.<sup>4</sup>: **B 01 J 8/24, C 08 F 10/00,**  
**C 08 F 2/34**

(54) Continuous process for the production of polymer in a fluidized bed reactor.

(30) Priority: 24.03.82 US 361547

(43) Date of publication of application:  
28.09.83 Bulletin 83/39

(45) Publication of the grant of the patent:  
15.11.89 Bulletin 89/46

(84) Designated Contracting States:  
AT BE CH DE FR GB IT LI NL SE

(50) References cited:  
EP-A-0 012 148  
EP-A-0 067 359  
FR-A-2 003 816  
US-A-2 751 756

(70) Proprietor: **UNION CARBIDE CORPORATION**  
**39 Old Ridgebury Road**  
**Danbury Connecticut 06817 (US)**

(72) Inventor: **Jenkins, John Mitchell, III**  
**1405 Village Drive**  
**South Charleston West Virginia 25309 (US)**  
Inventor: **Jones, Russell Lawrence**  
**7000 Falconbridge Road**  
**Chapel Hill North Carolina 27514 (US)**  
Inventor: **Jones, Thomas Michael**  
**1099 Hollyberry Lane**  
**South Charleston West Virginia 25309 (US)**  
Inventor: **Beret, Samil**  
**38 Maraposa Court**  
**Danville California 94526 (US)**

(74) Representative: **Weinhold, Peter, Dr. et al**  
**Patentanwälte Dr. V. Schmied-Kowarzik Dipl.-**  
**Ing. G. Dannenberg Dr. P. Weinhold Dr. D. Gudel**  
**Dipl.-Ing. S. Schubert Dr. P. Barz**  
**Siegfriedstrasse 8**  
**D-8000 München 40 (DE)**

Note: Within nine months from the publication of the mention of the grant of the European patent, any person may give notice to the European Patent Office of opposition to the European patent granted. Notice of opposition shall be filed in a written reasoned statement. It shall not be deemed to have been filed until the opposition fee has been paid. (Art. 99(1) European patent convention).

Courier Press, Leamington Spa, England.

**EP 0 089 691 B1**

## Description

This invention relates to a novel method for producing polymers improving the space time yield of an exothermic polymerization reaction conducted in a fluidized bed reactor.

5 The discovery of the fluidized bed process for the production of polymers provided a means for producing these diverse and widely used polymers with a drastic reduction in capital investment and a dramatic reduction in energy requirements as compared to then conventional processes. The present invention provides a means for even greater savings in energy and capital cost by affording a simple and efficient means for obtaining a substantial increase in production rate in a given size reactor over what was  
10 previously possible in a fluidized bed process.

The most common and perhaps universal means of heat removal employed in conventional fluidized bed reactor processes is by compression and cooling of the recycle gas stream at a point external to the reactor. In commercial scale fluidized bed reaction systems for producing polymers such as polyethylene, the amount of fluid which must be circulated to remove the heat of polymerization is greater than the  
15 amount of fluid required for support of the fluidized bed and for adequate solids mixing in the fluidized bed. The fluid velocity in the reactor is limited to prevent excessive entrainment of solids. A constant bed temperature will result if the heat generated by the polymerization reaction (which is proportional to the polymer production rate) is equal to the heat absorbed by the fluidizing stream as it passes through the bed, plus any heat removed or lost by other means.

20 Unfortunately, it has long been believed that the recycle gas temperature could not be lowered any further than to a point slightly above the dew point of the recycle gas stream. The dew point is that temperature at which liquid condensate begins to form in the gas stream. Common practice has been to limit the temperature of the recycle stream at the outlet of the cycle heat exchange zone to a temperature at least about 3 to 10°C above its dew point (see copending U.S. patent application Ser. No. 49,555 of June 18, 1979, equivalent to published European Patent Specification No. 0 021 605, page 22, lines 8—22). This  
25 assumption was predicated on the belief that the introduction of liquid into a gas phase fluidized bed reactor would inevitably result in plugging of the distribution plate, if one is employed; non-uniformity of monomer concentrations inside the fluidized bed and accumulation of liquid at the bottom of the reactor which would interfere with continuous operation or result in complete reactor shut-down. For products, such as those using hexene as a comonomer, the relatively high dew point of the recycle stream has until  
30 now severely restricted the production rate.

The primary limitation on reaction rate in a fluidized bed reactor is the rate at which heat can be removed from the polymerization zone. Although they differ in very important ways from gas fluidized bed reaction systems, the same heat limitation problems exist in other types of reaction systems such as stirred  
35 reaction systems and to some extent, slurry reaction systems.

In U.S. Patent No. 3,256,263, heat removal in a stirred reaction system is achieved by the compression of recycle gases and expansion upon reentry into the reactor. In other stirred or paddle-type reaction systems some additional cooling is effected by the injection of liquid onto the top of the bed. See for example U.S. Patents Nos. 3,254,070; 3,300,457 and 3,652,527.

40 In U.S. Patents Nos. 3,965,083; 3,970,611 and 3,971,768 assigned to Standard Oil Co., cooling of a stirred bed reactor is supplemented by injection of liquids on the top of the bed.

In U.S. Patent No. 4,012,573 (Treschman et al.) gases withdrawn from a stirred reactor are condensed to liquid and returned in liquid form to the stirred reactor where the liquid is brought into desired contact with polymer in the stirred bed.

45 Mitsubishi Petrochemical Co. has proposed the use of liquids or regasified liquids for cooling in a gas phase reactor (J55/045,744/80 and DT 2 139 182). In both of these descriptions the liquid or regasified liquid is injected into the bed rather than entering with the fluidizing gas as in the present invention. DT 2 139 182 is specific to stirred beds rather than fluidized beds. In J55/045,744/80 the liquid is regasified before being injected into the fluidized bed.

50 In a fluidized bed reaction system, as distinguished from stirred or paddle-type reaction systems, uniform distribution of monomer and catalysts in the upwardly moving gas stream is essential to avoid hot spots and resulting polymer chunks. In stirred and paddle-type reactors these problems are overcome by mechanical stirring and agitation. A further requirement of a fluidized bed reactor system is that the velocity of gas flowing through the reactor be adequate to maintain the bed in a fluidized state. The gas  
55 velocity required to keep the bed in a fluidized suspension cannot be achieved under normal conditions by mere injection of liquid at the bottom of the bed. Therefore, the direct liquid injection cooling of a reactor, as described by Treschman et al. is not a viable option for a fluidized bed reaction system.

Subject matter of the present invention is a continuous process for the production of polymer in a fluidized bed reactor from one or more fluid monomers by continuously passing a gaseous stream through  
60 said reactor in the presence of catalyst under reactive conditions, withdrawing polymeric product and unreacted fluid, cooling said unreacted fluids and returning said cooled fluids into said reactor together with sufficient additional monomers to replace those monomers polymerized and withdrawn as product, cooling part or all of said unreacted fluids to form a two-phase mixture of gas and entrained liquid below the dew point and reintroducing said two-phase mixture into said reactor.

65 While not limited to any particular type or kind of polymerization reaction (so long as the reaction is of

an exothermic nature), this invention is particularly suited to polymerization reactions involving the polymerization of one or more of the monomers listed below:

- I. Olefin type: ethylene, propylene, butene-1, pentene-1, 4-methylpentene-1, hexene-1, styrene.
  - II. Polar vinyl monomer type: vinyl chloride, vinyl acetate, vinyl acrylate, methyl methacrylate, tetrafluoroethylene, vinyl ether, acrylonitrile.
  - III. Diene type (conjugated and non-conjugated): butadiene, 1,4-hexadiene, isoprene, ethylidene norbornene.
  - IV. Acetylene type: acetylene, substituted acetylene, such as methyl acetylene.
  - V. Aldehyde type: formaldehyde.
- It is to be noted that catalysts employable in the fluidized bed polymerization of the above monomer types would, respectively, most usually be as follows:
- I. Coordinated anionic catalyst
  - II. Cationic catalyst for copolymers with ethylene only; others of this type require a free-radical catalyst
  - III. Either a free-radical catalyst or a coordinated anionic catalyst
  - IV. A coordinated anionic catalyst
  - V. An anionic catalyst

Although this invention is not limited to any specific type of polymerization reaction, the following discussions of the operation of the process are directed to a special process of polymerization of olefin-type monomers where the invention has been found to be especially advantageous.

In very general terms, a conventional fluidized bed process for producing resins, particularly polymers produced from monomers, is practiced by passing a gaseous stream containing one or more monomers continuously through a fluidized bed reactor under reactive conditions and in the presence of catalyst. The gaseous stream containing unreacted gaseous monomer is withdrawn from the reactor continuously, compressed, cooled and recycled into the reactor. Product is withdrawn from the reactor. Make-up monomer is added to the recycle stream.

The polymer-forming reaction is exothermic, making it necessary to maintain in some fashion the temperature of the gas stream inside the reactor at a temperature not only below the resin and catalyst degradation temperatures, but at a temperature below the fusion or sticking temperature of resin particles produced during the polymerization reaction. This is necessary to prevent plugging of the reactor due to rapid growth of polymer chunks which cannot be removed in a continuous fashion as product. It will be understood, therefore, that the amount of polymer that can be produced in a fluidized bed reactor of a given size in a specified time period is directly related to the amount of heat which can be withdrawn from the fluidized bed.

In accordance with this invention the recycle gas stream is intentionally cooled to a temperature below the dew point of the recycle gas stream to produce a two-phase gas-liquid mixture under conditions such that the liquid phase of said mixture will remain entrained in the gas phase of said mixture at least from the point of entry into the fluidized bed reactor until volatilized or until passage into the fluidized bed. A substantial increase in space time yield results from the practice of this invention with little or no change in product properties or quality. When practiced as described herein the overall process proceeds continuously and smoothly and without unusual operational difficulties.

It may be desirable in some instances to raise the dew point of the recycle gas stream to further increase heat removal. The recycle stream dew point can be increased by: (1) raising the operating pressure of the reaction system; (2) increasing the concentration of condensible fluids in the recycle stream; and/or (3) reducing the concentration of non-condensable gases in the recycle stream. In one embodiment of this invention, the dew point of the recycle stream may be increased by the addition of a condensible fluid to the recycle stream which is inert to the catalyst, reactants, and the products of the polymerization reaction. The fluid can be introduced into the recycle stream with the make-up fluid or by any other means or at any other point in the system. Examples of such fluids are saturated hydrocarbons, such as butanes, pentanes or hexanes.

A primary limitation on the extent to which the recycle gas stream can be cooled below the dew point is in the requirement that gas-to-liquid ratio be maintained at a level sufficient to keep the liquid phase of the two-phase fluid mixture in an entrained or suspended condition until the liquid is vaporized. It is also necessary that the velocity of the upwardly flowing fluid stream be sufficient to maintain the fluidized bed in a suspended condition.

While the liquid content of the two-phase recycle stream can be quite high, as general rule, the quantity of condensed liquid contained in the gas phase should not exceed about 20 weight percent and preferably should not exceed about 10 weight percent, provided always that the velocity of the two-phase recycle stream is high enough to keep the liquid phase in suspension in the gas and to support the fluidized bed within the reactor.

The entry point for the two-phase recycle stream should be below the fluidized bed (polymerization zone) to ensure uniformity of the upwardly flowing gas stream and to maintain the bed in a suspended condition. The recycle stream containing entrained liquid is introduced into the reactor at a point in the lower region of the reactor and most preferably at the very bottom of the reactor to ensure uniformity of the fluid stream passing upwardly through the fluidized bed.

A baffle or similar means for preventing regions of low gas velocity in the vicinity of the recycle stream entry point may be provided to keep solids and liquids entrained in the upwardly flowing recycle stream.

The two-phase mixture can be divided into at least two separate streams at least one of which is introduced into the reactor, e.g. at a point below the fluidized bed and one or more of said streams is optionally introduced into said fluidized bed, provided that the velocity of all gases introduced below said fluidized bed is at least sufficient to support said bed and to maintain it in a fluidized condition. In all cases the composition of the gas stream is kept essentially uniform and flowing in a manner such that there are no dead spaces in the bed where unremovable solids can form.

It will be apparent that if desired, it is possible to form a two-phase fluid stream within the reactor at the point of injection by separately injecting gas and liquid under conditions which will produce a two-phase stream. Little advantage is seen in operating in this fashion due to the added and unnecessary burden and cost of separating the gas and liquid phases after cooling. It may however, be desirable to inject make-up monomer into the reactor in this fashion. The injection of liquid or gaseous make-up monomer at the point of entry of the two-phase recycle stream or elsewhere in the reactor or in the recycle stream is contemplated by this invention. Additional quantities of one or more monomers can be added to replace monomers polymerized and withdrawn as product are introduced into the recycle stream prior to entry of the said two-phase mixture into the reactor.

The advantages of this invention are not limited to the production of polyolefin resins. This invention can be practiced in connection with any exothermic polymerization process carried out in a gas phase fluidized bed. The advantages of this invention over conventional processes will generally increase in direct relation to the nearness of the dew point temperature of the recycle stream to the reaction temperature within the interior of the fluid bed. For obvious reasons this invention cannot be used in a reaction system in which the reaction temperature within its fluidized bed is below the dew point of the recycle gas stream.

The applicability of this invention to the production of any given polymer can be determined by the use of the following formula:

$$X = \frac{P \cdot H_{rxn}}{G_{mass} \cdot CP_{gas} (T_{rxn} - T_{limit})}$$

P=desired polymer production rate: constrained to rates giving X less than 1.0 without subject invention.

$H_{rxn}$ =heat of polymerization of specific polymer being produced.

$G_{mass}$ =mass flow rate of recycle stream; limited to a minimum value by the need for adequate fluidization and mixing in the bed and to a maximum value by entrainment of solids. Specific minima and maxima depend on numerous factors known to those skilled in the art.

$CP_{gas}$ =heat capacity of the recycle stream.

$T_{rxn}$ =temperature of the reaction zone (fluid bed); has maximum value depending on the sticking temperature of the polymer at the pressure of the recycle stream and/or the catalyst performance, and a minimum value which is dependent on catalyst performance.

$T_{limit}$ =minimum temperature of the recycle stream entering the reaction zone as limited prior to the present invention. This temperature is either the dew point of the recycle stream or the cooling limit of the heat exchange zone, whichever is higher. If  $T_{limit}$  is the recycle stream dew point, the invention is practiced by simply cooling the stream to a temperature below its dew point. If  $T_{limit}$  is controlled by the heat exchange zone, the invention is practiced by adding a condensable fluid to increase the dew point of the recycle stream to a temperature above the cooling limit of the heat exchange zone.

Whenever the value of X is greater than 1, the use of this invention will afford a benefit and as the value of X increases, the greater are the benefits which can result from this invention.

A fluidized bed reaction system which is particularly suited to production of polyolefin resin by the practice of the process of the present invention is illustrated in the drawing. With reference thereto, the reactor 10 consists of a reaction zone 12 and a velocity reduction zone 14.

In general, the height to diameter ratio of the reaction zone can vary in the range of about 2.7:1 to about 4.6:1. The range, of course, can vary to larger or smaller ratios and depends upon the desired production capacity. The cross-sectional area of the velocity reduction zone 14 is typically within the range of about 2.6 to about 2.8 multiplied by the cross-sectional area of the reaction zone 12.

The reaction zone 12 includes a bed of growing polymer particles, formed polymer particles and a minor amount of catalyst particles fluidized by the continuous flow of polymerizable and modifying gaseous components in the form of make-up feed and recycle fluid through the reaction zone. To maintain a viable fluidized bed, the superficial gas velocity through the bed must exceed the minimum flow required for fluidization, and preferably is at least 6 cm/s (0.2 ft/sec) above minimum flow. Ordinarily, the superficial gas velocity does not exceed 1.5 m/s (5.0 ft/sec) and usually no more than 0.76 m/s (2.5 ft/sec) is sufficient.

It is essential that the bed always contain particles to prevent the formation of localized "hot spots," and to entrap and distribute the particulate catalyst throughout the reaction zone. On start up, the reactor is usually charged with a base of particulate polymer particles before gas flow is initiated. Such particles may be identical in nature to the polymer to be formed or different therefrom. When different, they are

withdrawn with the desired formed polymer particles as the first product. Eventually, a fluidized bed of desired polymer particles supplants the start-up bed.

The partially or totally activated precursor composition and/or catalyst used in the fluidized bed is preferably stored for service in a reservoir 16 under a blanket of a gas which is inert to the stored material, such as nitrogen or argon.

Fluidization is achieved by a high rate of fluid recycle to and through the bed, typically in the order of about 50 times the rate of feed of make-up fluid. The fluidized bed has the general appearance of a dense mass of individually moving particles as created by the percolation of gas through the bed. The pressure drop through the bed is equal to or slightly greater than the weight of the bed divided by the cross-sectional area. It is thus dependent on the geometry of the reactor.

Make-up fluid is fed to the recycle line at point 18. The composition of the make-up stream is determined by a gas analyzer 21. The gas analyzer determines the composition of the recycle stream and the composition of the make-up stream is adjusted accordingly to maintain an essentially steady state gaseous composition within the reaction zone.

The gas analyzer is a conventional gas analyzer which operates in conventional manner to indicate recycle stream composition and which is adapted to regulate the feed and is commercially available from a wide variety of sources. Generally, the gas analyzer 21 can be positioned so as to receive gas from a point between the velocity reduction zone 14 and heat exchanger 24.

To ensure complete fluidization, the recycle stream and, where desired, part of the make-up stream are returned through recycle line 22 to the reactor at point 26 below the bed. There may be preferably a gas distributor plate 28 above the point of return to aid in fluidizing the bed. In passing through the bed, the recycle stream absorbs the heat of reaction generated by the polymerization reaction.

The portion of the fluidizing stream which does not react in the bed constitutes the recycle stream which is removed from the polymerization zone, preferably by passing it into velocity reduction zone 14 above the bed where entrained particles are given an opportunity to drop back into the bed.

The recycle stream is then compressed in a compressor 30 and then passed through a heat exchange zone wherein the heat of reaction is removed before it is returned to the bed. The heat exchange zone is typically a heat exchanger 24 which can be of the horizontal or vertical type. The recycle stream is then returned to the reactor at its base 26 and to the fluidized bed through gas distributor plate 28. A gas deflector 32 is preferably installed at the inlet to the reactor to prevent contained polymer particles from settling out and agglomerating into a solid mass.

The temperature of the bed is controlled at an essentially constant temperature under steady state conditions by constantly removing the heat of reaction. No noticeable temperature gradient appears to exist within the upper portion of the bed. A temperature gradient will exist in the bottom of the bed in a layer of about 15 to 30 cm (6 to 12 inches), between the temperature of the inlet fluid and the temperature of the remainder of the bed.

Good gas distribution plays an important role in the operation of the reactor. The fluidized bed contains growing and formed particulate polymer particles, as well as catalyst particles. As the polymer particles are hot and possibly active, they must be prevented from settling, for if a quiescent mass is allowed to exist, any active catalyst contained therein may continue to react and cause fusion. Diffusing recycle fluid through the bed at a rate sufficient to maintain fluidization through the bed is, therefore, important.

Gas distribution plate 28 is a preferred means for achieving good gas distribution and may be a screen, slotted plate, perforated plate, a plate of the bubble-cap type and the like. The elements of the plate may all be stationary, or the plate may be of the mobile type disclosed in U.S. 3,298,792. Whatever its design, it must diffuse the recycle fluid through the particles at the base of the bed to keep the bed in a fluidized condition, and also serve to support a quiescent bed of resin particles when the reactor is not in operation.

The preferred type gas distributor plate 28 is generally of the type which is fabricated from metal and which has holes distributed across its surface. The holes are normally of a diameter of about 1.3 cm (1/2 inch). The holes extend through the plate, and over each hole there is positioned a triangular angle iron identified as reference numeral 36 which is fixedly mounted to plate 28. The angle irons serve to distribute the flow of fluid along the surface of the plate so as to avoid stagnant zones of solids. In addition they prevent the resin from flowing through the holes when the bed is settled.

Any fluid inert to the catalyst and reactants can also be present in the recycle stream. An activator compound, if utilized, is preferably added to the reaction system downstream from heat exchanger 24. Thus, the activator may be fed into the recycle stream from dispenser 38 through line 40.

It is essential to operate the fluid-bed reactor at a temperature below the sintering temperature of the polymer particles to ensure that sintering will not occur. The sintering temperature is a function of resin density. In general, polyethylene low-density resins, for example, have a low sintering temperature and polyethylene high-density resins, for example, have a higher sintering temperature. For example, temperatures of from about 75°C to about 95°C are used to prepare ethylene copolymers having a density of from about 0.91 g/cm<sup>3</sup> to about 0.95 g/cm<sup>3</sup>, while temperatures of from about 100°C to about 115°C are used to prepare ethylene copolymers or homopolymers having a density of from about 0.95 g/cm<sup>3</sup> to about 0.97 g/cm<sup>3</sup>.

The fluid-bed reactor may be operated at pressures of up to about 6.9 MPa (1000 psi), and is for polyolefin resin production preferably operated at a pressure of from about 0.69 MPa (100 psi) to about 2.4

MPa (350 psi), with operation at the higher pressures in such ranges favoring heat transfer since an increase in pressure increases the unit volume heat capacity of the gas.

The partially or totally activated precursor composition and/or catalyst (hereinafter collectively referred to as catalyst) is injected into the bed at a rate equal to its consumption at a point 42 which is above distributor plate 28. Preferably, the catalyst is injected at a point in the bed where good mixing of polymer particles occurs. Injecting the catalyst at a point above the distribution plate is an important feature for satisfactory operation of a fluidized bed polymerization reactor. Since catalysts are highly active, injection of the catalyst into the area below the distributor plate may cause polymerization to begin there and eventually cause plugging of the distributor plate. Injection into the fluidized bed, instead, aids in distributing the catalyst throughout the bed and tends to preclude the formation of localized spots of high catalyst concentration which may result in the formation of "hot spots". Injection of the catalyst into the reactor above the bed may result in excessive catalyst carryover into the recycle line where polymerization may begin and plugging of the line and heat exchanger may eventually occur.

The catalyst can be injected into the reactor by various techniques. It is preferred, however, to continuously feed the catalyst into the reactor utilizing a catalyst feeder as disclosed; e.g., in U.S. patent 3,779,712. The catalyst is preferably fed into the reactor at a point 20 to 40 percent of the reactor diameter away from the reactor wall and at a height of about 5 to about 30 percent of the height of the bed.

A gas which is inert to the catalyst, such as nitrogen or argon, is preferably used to carry the catalyst into the bed.

The rate of polymer production in the bed depends on the rate of catalyst injection and the concentration of monomer(s) in the recycle stream. The production rate is conveniently controlled by simply adjusting the rate of catalyst injection.

Since any change in the rate of catalyst injection will change the reaction rate and hence rate of generation of the heat of reaction, the temperature of the recycle stream entering the reactor is adjusted upwards and downwards to accommodate any change in the rate of heat generation. This ensures the maintenance of an essentially constant temperature in the bed. Complete instrumentation of both the fluidized bed and the recycle stream cooling system is of course, useful to detect any temperature change in the bed so as to enable either the operator or a conventional automatic control system to make a suitable adjustment in the temperature of the recycle stream.

Under a given set of operating conditions, the fluidized bed is maintained at essentially a constant height by withdrawing a portion of the bed as product at the rate of formation of the particulate polymer product. Since the rate of heat generation is directly related to the rate of product formation, a measurement of the temperature rise of the fluid across the reactor (the difference between inlet fluid temperature and exit fluid temperature) is indicative of the rate of particulate polymer formation at a constant fluid velocity if no vaporizable liquid is present in the inlet fluid.

On discharge of particulate polymer product from reactor 10, it is desirable and preferable to separate fluid from the product and to return the fluid to the recycle line 22. There are numerous ways known in the art to accomplish this. One preferred system is shown in the drawings. Thus, fluid and product leave reactor 10 at point 44 and enter product discharge tank 46 through valve 48 which is designed to have minimum restriction to flow when opened such as a ball valve. Positioned above and below product discharge tank 46 are conventional valves 50, 52 with the latter being adapted to provide passage of product into product surge tank 54. Product surge tank 54 has venting means illustrated by line 56 and gas entry means illustrated by line 58. Also positioned at the base of product surge tank 54, is a discharge valve 60 which when in the open position discharges product for conveying to storage. Valve 50 when in the open position releases fluid to surge tank 62. Fluid from surge tank 62 is directed through a filter absorber 64 and thence through a compressor 66 and into recycle line 22 through line 68.

In a typical mode of operation, valve 48 is open and valves 50, 52 are in a closed position. Product and fluid enter product discharge tank 46. Valve 48 closes and the product is allowed to settle in product discharge tank 46. Valve 50 is then opened permitting fluid to flow from product discharge tank 46 to surge tank 62 from which it is continually compressed back into recycle line 22. Valve 50 is then closed and valve 52 is opened and any product in product discharge tank 46 flows into product surge tank 54. Valve 52 is then closed. The product is purged with inert gas, preferably nitrogen, which enters product surge tank 54 through line 58 and is vented through line 56. Product is then discharged from product surge tank 54 through valve 60 and conveyed through line 20 to storage.

The particular timing sequence of the valves is accomplished by the use of conventional programmable controllers which are well known in the art. Moreover, the valves can be kept substantially free of agglomerated particles by installation of provisions for directing a stream of gas periodically through the valves and back to the reactor.

Another preferred product discharge system which may be alternatively employed is that disclosed and claimed in the copending U.S. Patent application of Robert G. Aronson filed July 28, 1981, Ser. No. 287815 (EP—A—0 071 430 publ. 9.2.83) and entitled Fluidized Bed Discharge System. Such a system employs at least one (parallel) pair of tanks comprising a settling tank and a transfer tank arranged in series and having the separated gas phase returned from the top of the settling tank to a point in the reactor near the top of the fluidized bed. Such alternative preferred product discharge system obviates the need for a recompression line 64, 66, 68, as shown in the system of the drawing.

The fluidized-bed reactor is equipped with an adequate venting system (not shown) to allow venting the bed during start up and shut down. The reactor does not require the use of stirring and/or wall scraping. The recycle line 22 and the elements therein (compressor 30, heat exchanger 24) should be smooth surfaced and devoid of unnecessary obstructions so as not to impede the flow of recycle fluid or entrained particles.

Among the polymers which may be produced in the process of the present invention are homopolymers of ethylene, propylene, butene or copolymers of a major mole percent of ethylene, propylene or butene and a minor mole percent of one or more  $C_2$  to  $C_8$  alpha-olefins. The  $C_2$  to  $C_8$  alpha-olefins preferably should not contain any branching on any of their carbon atoms which is closer than the fourth carbon atom. The preferred  $C_2$  to  $C_8$  alpha-olefins are ethylene, propylene, butene-1, pentene-1, hexene-1, 4-methylpentene-1 and octene-1.

The ethylene polymers, for example, have a melt flow ratio of over about 22. The melt flow ratio value is another means of indicating the molecular weight distribution of a polymer. A melt flow ratio (MFR) of 22 thus, for example, corresponds to a Mw/Mn value (as determined by conventional size exclusion chromatography) of about 2.7.

The ethylene homopolymers have a density of about 0.958 to 0.972 g/cm<sup>3</sup>.

The ethylene copolymers have a density less than about 0.96 g/cm<sup>3</sup>. The density of the ethylene copolymer, at a given melt index level for the copolymer, is primarily regulated by the amount of the  $C_3$  to  $C_8$  comonomer which is copolymerized with the ethylene. In the absence of the comonomer, the ethylene would homopolymerize to provide polymers having a density of about  $\leq 0.96$ . Thus, the addition of progressively larger amounts of the comonomers to the copolymers results in a progressive lowering of the density of the copolymer. The amount of each of the various  $C_3$  to  $C_8$  comonomers needed to achieve the same result will vary from monomer to monomer, under the same reaction conditions.

Thus, to produce binary copolymers of ethylene with the same density and melt index, larger molar amounts of the different comonomers would be needed in the order of  $C_3 > C_4 > C_5 > C_6 > C_7 > C_8$ .

When made in the fluid-bed process described herein, ethylene polymers are granular materials which have a settled bulk density of about 0.24 to 0.51 g/cm<sup>3</sup> (15 to 32 pounds per cubic foot) and an average particle size of the order of about 0.13 (0.005) to about 1.52 mm (0.06 inches). Particle size is important for the purposes of readily fluidizing the polymer particles in the fluid bed reactor, as herein described.

In accordance with the present invention, a process is provided for increasing the space time yield of polymer production in a fluidized bed reactor employing an exothermic polymerization reaction by cooling the recycle stream to below its dew point and returning the resultant two-phase fluid stream to the reactor to maintain the fluidized bed at a desired temperature above the dew point of the recycle stream. The cooling capacity of the recycle stream is increased both due to the greater temperature differential between the entering recycle stream and the reactor and by the vaporization of the condensed liquids entrained in the recycle stream.

It has, therefore, been found that the dew point limitation on the recycle stream broadly held heretofore by those skilled in the art, as pointed out hereinabove, has been unduly limiting and restricting. It has been found that such limitations are erroneous and need not be practiced.

The amount of condensation, and thus the increase in production rate, can be further enhanced by altering the process conditions so as to increase the dewpoint of the recycle stream. It has been found that the amount of condensation of liquid in the recycle stream can be maintained at up to about 20 percent by weight, preferably up to about 10 percent. This degree of condensation is achieved by maintaining the outlet temperature from the cycle heat exchange zone so as to effect the required degree of cooling below the dew point of the mixture.

Subject matter of the divisional application 87 106 294.9 (publication number EP—A—241947) is a method for controlling the temperature of a fluidized bed during the production of polymers in a fluidized bed reactor by an exothermic polymerization reaction, which comprises continuously introducing into the bed a stream of gas cooled to below the maximum desired temperature within said bed and simultaneously or separately introducing a stream of liquid into said reactor under conditions such that an essentially uniform two-phase mixture of said gas and liquid is introduced in said bed at a level below the region of maximum desired temperature within said reactor. This means that a more general aspect of the present invention is claimed there.

A further subject matter of the divisional application mentioned above is a process for producing polymers from one or more monomers by an exothermic polymerization reaction in a fluidized bed reactor having an upper polymerization zone containing a bed of growing polymer particles and a lower gas diffusion zone, which comprises:

(1) continuously introducing a two-phase gas-liquid stream containing one or more monomers into said polymerization zone with an upward velocity sufficient to maintain said particles in a suspended and gas fluidized condition;

(2) continuously introducing a polymerization catalyst into said polymerization zone;

(3) continuously withdrawing polymer product from said polymerization zone; and

(4) continuously withdrawing unreacted gases from the said polymerization zone, compressing and cooling said gases to a temperature below the dew point of said gases.

Also this process in some respects more general since any two-phase gas-liquid stream can be fed into the polymerization zone.

#### Example 1

5 In an example of the process of the invention, a fluidized bed gas phase reaction system, as described above was operated with a partially condensed recycle stream to produce an hexene-1/propylene/ethylene copolymer, under the following conditions:

Recycle Stream Composition (mole percent):

10 Nitrogen—34.7  
Ethylene—38.1  
Ethane—8.9  
Propylene—5.1  
Hexene-1—4.8  
Hydrogen—7.3  
15 Methane—0.1

Recycle Stream Dewpoint: 57.8°C (at the pressure of the heat exchanger)

Reactor Gas Inlet Temperature: 48°C

Liquid in Cycle Gas: 4.6 wt. percent

20 Catalyst: a complex of tetrahydrofuran, magnesium chloride and titanium chloride reduced with diethyl aluminum chloride (diethyl aluminum chloride-to-tetrahydrofuran molar ratio of 0.4) and tri-n-hexyl aluminum (tri-n-hexyl aluminium-to-tetrahydrofuran molar ratio of 0.4) impregnated on triethyl aluminum treated silica dioxide.

Reactor Temperature: 88°C

Reactor Pressure: 1963 kPa (270 psig)

25 Superficial Gas Velocity in the Fluidized Bed: 0.61 m/s (2.0 feet/sec)

Bed Height: 11.3 m (37 feet)

Bed Diameter: 2.4 m (8 feet)

Space Time Yield: 81.7 kg/m<sup>3</sup>h (5.1 lb/ft<sup>3</sup>h)

Melt Index of Resin: 1.0 dg/min

30 Density of Resin: 0.918 g/cm<sup>3</sup>

Activator: triethyl aluminum

#### Example 1a

35 The following example was carried out under essentially identical operating conditions as Example 1, differing only with respect to the cooling of the recycle stream and the rate of catalyst injection.

Recycle Stream Composition (mole percent):

40 Nitrogen—36.5  
Ethylene—38.2  
Ethane—11.2  
Propylene—5.0  
Hexene-1—4.1  
Hydrogen—4.8

Methane—None detected

Recycle Stream Dew Point: 53.2°C (at the pressure of the heat exchanger)

45 Reactor Gas Inlet Temperature: 60.6°C

Liquid in Cycle Gas: Zero

Reactor Temperature: 87.1°C

Reactor Pressure: 1963 kPa (270 psig)

Superficial Gas Velocity in the Fluidized Bed: 0.61 m/s (2.0 ft/sec)

50 Bed Height: 11.3 m (37 feet)

Bed Diameter: 2.4 m (8 feet)

Space Time Yield: 32 kg/m<sup>3</sup>h (2.0 lb/ft<sup>3</sup>hr)

Melt Index of Resin: 1.0 dg/min

Density of Resin: 0.918 g/cm<sup>3</sup>

55 Activator: triethyl aluminum

It is to be noted that operation below the dew point of the recycle stream, as in Example 1, resulted in an increase in space time yield (pounds of resin produced per cubic foot of bed per hour) which was approximately 2.5 times those obtained without the employment of the liquid phase in the recycle stream, as in this Example 1a.

60

65



## Example 2

In another Example, employing butene-1 rather than propylene and hexene-1 as the other alpha-olefin monomer, the following conditions were maintained:

- Recycle Stream Composition (mole percent)
- Hydrogen—6.3
  - Ethylene—57.8
  - Butene-1—26.7
  - Inerts—9.2
- Recycle Stream Dewpoint: 49.7°C (at the pressure of the heat exchanger)
- Reactor Gas Inlet Temperature: 47.2°C
- Liquid in Cycle Gas: 1.2 wt. percent
- Catalyst: a complex of tetrahydrofuran, magnesium chloride and titanium chloride reduced with tri-n-hexyl aluminum only (tri-n-hexyl aluminum-to-tetrahydrofuran molar ratio of 0.7) impregnated on tri-ethyl aluminum treated silica dioxide.
- Reactor Temperature: 87.5°C
  - Reactor Pressure: 1956 kPa (269 psig)
  - Superficial Gas Velocity in the Fluidized Bed: 0.61 m/s (2.0 ft/sec)
  - Bed Height: 11.3 m (37 feet)
  - Bed Diameter: 24 m (8 feet)
  - Space Time Yield: 99 kg/m<sup>3</sup>h (6.2 lb/ft<sup>3</sup>hr)
  - Melt Index of Resin: 1.0 dg/min
  - Density of Resin: 0.918 g/cm<sup>3</sup>
  - Activator: triethyl aluminum

## Example 2a

The following example was conducted under conditions similar to those employed in example 2 but without condensed liquid in the recycle gas stream.

- Recycle Stream Composition (mole percent):
- Hydrogen—4.5
  - Ethylene—59.3
  - Butene-1—26.7
  - Inerts—9.5
- Recycle Stream Dew Point: 44.1°C (at the pressure of the heat exchanged)
- Reactor Gas Inlet Temperature: 56°C
- Liquid in Cycle Gas: Zero
- Catalyst: A complex of tetrahydrofuran, magnesium chloride and titanium chloride reduced with tri-n-hexyl aluminum only (tri-n-hexyl aluminum-to-tetrahydrofuran molar ratio of 0.7) impregnated on tri-ethyl aluminum treated silica dioxide.
- Reactor Temperature: 83.7°C
  - Reactor Pressure: 1928 kPa (265 psig)
  - Superficial Gas Velocity in the Fluidized Bed: 0.65 m/s (2.14 ft/sec)
  - Bed Height: 11.4 m (37.5 feet)
  - Bed Diameter: 2.4 m (8 feet)
  - Space Time Yield: 59 kg/m<sup>3</sup>h (3.7 lb/ft<sup>3</sup>hr)
  - Melt Index of Resin: 1.0 dg/min
  - Density of Resin: 0.918 g/cm<sup>3</sup>
  - Activator: triethyl aluminum

## Examples 3—8

- The six following examples in tabular form set forth further information for practicing the process of the invention employing various condensing recycle streams for enhanced cooling of the fluidized bed in the polymerization of hexene-1/butene-1/ethylene, ethylene/propylene, ethylene/butene-1, pentene-1/ethylene, as well as the use of iso-pentane as an inert fluid in the polymerization of butene-1/ethylene and ethylene, respectively.

		Example No.		
		3	4	5
5	Product	C <sub>6</sub> /C <sub>4</sub> /C <sub>2</sub> Copolymer	C <sub>2</sub> /C <sub>3</sub> Copolymer	C <sub>2</sub> /C <sub>4</sub> Copolymer
	Recycle stream composition (mole percent):			
10	Nitrogen	34.7	17.4	43.5
	Ethylene	38.1	0.64	0.25
	Ethane	8.9	—	—
	Propylene	—	63.5	—
	Propane	—	16.5	—
15	Butene-1	5.1	—	50.0
	Butane	—	—	5.0
	Pentene-1	—	—	—
	Isopentane	1.0	1.0	0.5
	Hexene-1	4.8	—	—
20	Hydrogen	7.3	0.95	0.75
	Methane	0.1	—	—
	Recycle stream Dewpoint (°C)*	63.0	44.4	44.0
25	Reactor inlet temperature (°C)	54.3	43.5	41.6
	Liquid in cycle gas (wt. percent)	4.8	4.2	8.8
30	Reactor temperature (°C)	88	60	50
	Reactor pressure kPa (psig)	1963 (270)	1963 (270)	935 (121)
35	Superficial gas velocity in fluidized bed m/s (ft/sec)	0.61 (2.0)	0.52 (1.7)	0.61 (2.0)
40	Bed height m (feet)	11.3 (37)	11.3 (37)	11.3 (37)
45	Space time yield kg/m <sup>3</sup> h (lb/ft <sup>3</sup> hr.)	88 (5.5)	106 (6.6)	64 (4.0)

\* At the pressure of the heat exchanger

		Example No.		
		6	7	8
5	Product	C <sub>2</sub> /C <sub>2</sub> Copolymer	C <sub>2</sub> /C <sub>2</sub> Copolymer	C <sub>2</sub> Homopolymer
	Recycle stream composition (mole percent):			
10	Nitrogen	40.7	15.7	37.5
	Ethylene	35.0	38.6	35.0
	Ethane	9.0	3.0	5.0
	Propylene	—	—	—
	Propane	—	—	—
15	Butene-1	—	17.7	—
	Butane	—	2.3	—
	Pentene-1	10.5	—	—
	Isopentane	—	14.2	10.5
	Hexene-1	—	—	—
20	Hydrogen	3.8	6.2	11.0
	Methane	1.0	2.3	1.0
	Recycle stream dewpoint (°C)*	53.7	74.4	47.4
25	Reactor inlet temperature (°C)	42.5	65.9	34.0
	Liquid in cycle gas (wt. percent)	8.4	11.5	10.5
30	Reactor temperature (°C)	88	88	110
	Reactor pressure kPa (psig)	1963 (270)	1963 (270)	1963 (270)
35	Superficial gas velocity in fluidized bed m/s (ft/sec)	0.61 (2.0)	0.61 (2.0)	0.61 (2.0)
40	Bed height m (feet)	11.3 (37)	11.3 (37)	11.3 (37)
	Space time yield kg/m <sup>3</sup> h (lb/ft <sup>3</sup> hr.)	128 (8.0)	112 (7.0)	171 (10.7)

\* At the pressure of the heat exchanger

The space time yield improvements achieved by the practice of this invention are the result of the increased cooling capacity of the recycle stream. This increased capacity is due both to the greater temperature differential between the entering recycle stream and the bed temperature and to the evaporation of condensed liquid entrained in the recycle stream.

It will be appreciated that increased cooling is achieved not only by evaporation of entering entrained liquid, but also by the overall reduction in the temperature of both the gas and liquid phases of the recycle stream in comparison to previously known methods of operation of fluidized bed reactor systems.

Although the process of the invention can be applied to increase the production rate from existing and new commercial reactors, the invention also has the potential of reducing equipment cost in new reactors. For example, plants designed according to the prior art for the production of certain ethylene copolymers made using hexene-1 as an other alpha olefin must have larger diameter reactors due to the lower space time yield. Since this invention increases the space time yield for products made with hexene-1 to that of conventional butene-1/ethylene copolymers, the extra cost associated with the larger diameter reactor can be eliminated by practicing the present invention. Alternately, the large diameter reactor can be retained and the velocity of the recycle stream reduced, resulting in lower capital and operating costs for the cycle gas compressor.

Advantages of this invention over other heat removal methods which use condensed liquids include the simplicity of this invention, good liquid distribution, maximum heat removal, and uniform gas stream

composition in the polymerization zone as a result of introducing the two-phase recycle stream in the lower regions of the reactor. No liquid collection, separation, or injection equipment is required and the liquids are dispersed in the stream entering the fluidized bed over the entire cross-sectional area, avoiding temperature gradients and high local concentrations of liquid around injection means.

5

## Claims

1. A continuous process for the production of polymer in a fluidized bed reactor from one or more fluid monomers by continuously passing a gaseous stream through said reactor in the presence of catalyst  
10 under reactive conditions, withdrawing polymeric product and unreacted fluids, cooling said unreacted fluids and returning said cooled fluids into said reactor together with sufficient additional monomers to replace those monomers polymerized and withdrawn as product, which comprises: cooling part or all of said unreacted fluids to form a two-phase mixture of gas and entrained liquid below the dew point and reintroducing said two-phase mixture into said reactor.
- 15 2. A process according to claim 1 wherein said two-phase mixture of gas and entrained liquid is introduced into said reactor at a point below the fluidized bed.
3. A process according to claim 1 wherein said unreacted fluids are compressed prior to cooling.
4. A process according to claim 1 wherein means are provided to maintain the liquid phase of said two-phase mixture entrained in the gas phase of said mixture until such time as such liquids become  
20 volatilized or enter the fluidized bed of said reactor.
5. A process according to claim 1 wherein the velocity of said gas stream within said reactor is at or above a level sufficient to maintain said bed in a suspended and fluidized condition.
6. A process according to claim 1 wherein an inert condensible fluid is present in said two-phase mixture to raise the dew point of said mixture.
- 25 7. A process according to claim 6 wherein said inert condensible fluid is a saturated hydrocarbon.
8. A process according to claim 1 wherein the liquid phase of said two-phase mixture is less than about 20 percent by weight of the total weight of said two-phase mixture.
9. A process according to claim 1 wherein the liquid phase of said two-phase mixture is less than about 10 percent by weight of the total weight of said two-phase mixture.
- 30 10. A process according to claim 1 wherein the flow rate of said two-phase mixture and the ratio of liquid to gas in said mixture are maintained at a level sufficient to maintain said liquid entrained in said gas until volatilized or until said liquid enters the fluidized bed of said reactor.
11. A process according to claim 1 wherein said two-phase mixture is divided into at least two separate streams at least one of which is introduced into the reactor at a point below the fluidized bed and one or  
35 more of said streams is optionally introduced into said fluidized bed, provided that the velocity of all gases introduced below said fluidized bed is at least sufficient to support said bed and to maintain it in a fluidized condition.
12. A process according to claim 1 wherein the gas stream flowing upwardly through said fluidized bed is essentially uniform in composition and moving in a manner through said bed such that the entire bed is  
40 maintained in a fluidized condition and with essentially no dead spaces through which gases do not pass.
13. A process according to claim 1 wherein said additional monomers added to replace monomers polymerized and withdrawn as product, are introduced into the reactor in liquid form, below the fluidized bed and at or near the point of entry of said two-phase mixture into said reactor.
14. A process according to claim 1 wherein said additional monomers added to replace monomers  
45 polymerized and withdrawn as product are introduced into the recycle stream prior to entry of the said two-phase mixture into the reactor.
15. A process according to claim 1 wherein gas distributor means is provided immediately below the fluidized bed and above the region wherein said two-phase mixture is introduced into said reactor.
16. A process according to claim 1 wherein the internal pressure within said reactor is from about 0.69  
50 MPa (100 psi) to about 2.4 MPa (350 psi).
17. A process according to claim 1 wherein said monomer is an alpha olefin monomer having from two to four carbon atoms alone or in admixture with one or more other alpha olefin monomers having from two to eight carbon atoms.

55

## Patentansprüche

1. Kontinuierliches Verfahren zur Herstellung eines Polymers in einem Fließbettreaktor aus einem oder mehreren fließbaren Monomeren durch kontinuierliches Hindurchleiten eines gasförmigen Stromes durch  
60 diesen Reaktor unter Reaktionsbedingungen in Anwesenheit eines Katalysators, Abziehen des polymeren Produktes und der nicht-umgesetzten fließbaren Materialien, Abkühlen dieser nicht-umgesetzten fließbaren Materialien und Rückführung der abgekühlten fließbaren Materialien in den Reaktor zusammen mit ausreichenden zusätzlichen Monomeren, um die polymerisierten und als Produkt abgezogenen Monomeren zu ersetzen, wobei das Verfahren die Merkmale umfaßt, daß ein Teil oder alle nicht-  
65 umgesetzten fließbaren Materialien zur Bildung einer Zwei-Phasen-Mischung aus Gas und mitgeführter

Flüssigkeit unterhalb den Taupunkt abgekühlt und diese Zwei-Phasen-Mischung erneut in den Reaktor eingeführt wird.

2. Verfahren nach Anspruch 1, worin die Zwei-Phasen-Mischung aus Gas und mitgeführter Flüssigkeit an einem Punkt unterhalb des Fließbettes in den Reaktor eingeführt wird.

5 3. Verfahren nach Anspruch 1, worin die nicht-umgesetzten fließbaren Materialien vor dem Abkühlen komprimiert werden.

4. Verfahren nach Anspruch 1, worin Mittel vorgesehen sind, um die flüssige Phase dieser Zwei-Phasen-Mischung, die in der Gasphase dieser Mischung mitgeführt wird, bis zu einem Zeitpunkt aufrechtzuerhalten, wo diese Flüssigkeiten verdampft werden oder in das Fließbett des Reaktors eintreten.

10 5. Verfahren nach Anspruch 1, worin die Geschwindigkeit des Gasstromes innerhalb des Reaktors sich bei oder oberhalb eines ausreichenden Wertes befindet, um das Bett in suspendiertem und fließbarem Zustand zu halten.

6. Verfahren nach Anspruch 1, worin ein inertes kondensierbares fließbares Material in der Zwei-Phasen-Mischung anwesend ist, um den Taupunkt dieser Mischung zu erhöhen.

15 7. Verfahren nach Anspruch 6, worin das inerte kondensierbare fließbare Material ein gesättigter Kohlenwasserstoff ist.

8. Verfahren nach Anspruch 1, worin die flüssige Phase der Zwei-Phasen-Mischung weniger als etwa 20 Gew.-% des gesamten Gewichtes dieser Zwei-Phasen-Mischung ausmacht.

9. Verfahren nach Anspruch 1, worin die flüssige Phase der Zwei-Phasen-Mischung weniger als etwa 20 Gew.-% des gesamten Gewichtes dieser Zwei-Phasen-Mischung ausmacht.

20 10. Verfahren nach Anspruch 1, worin die Fließgeschwindigkeit der Zwei-Phasen-Mischung und das Verhältnis der Flüssigkeit zum Gas in dieser Mischung auf einem ausreichenden Wert gehalten wird, daß diese Flüssigkeit solange im Gas mitgeführt wird, bis sie verdampft oder bis die Flüssigkeit in das Fließbett des Reaktors eintritt.

25 11. Verfahren nach Anspruch 1, worin die Zwei-Phasen-Mischung in mindestens zwei getrennte Ströme geteilt wird, von denen mindestens einer an einem Punkt unterhalb des Fließbettes in den Reaktor eingeführt wird und einer oder mehrere dieser Ströme wahlweise in das Fließbett eingeführt wird bzw. werden, mit der Maßgabe, daß die Geschwindigkeit aller, unterhalb des Fließbettes eingeführten Gase mindestens ausreicht, um das Bett zu tragen und es in fließbarem Zustand zu halten.

30 12. Verfahren nach Anspruch 1, worin der aufwärts durch das Fließbett fließende Gasstrom von im wesentlichen einheitlicher Zusammensetzung ist und sich in solcher Weise durch das Bett bewegt, daß das gesamte Bett in fließbarem Zustand gehalten wird und praktisch keine toten Räume auftreten, durch welche die Gase nicht hindurchströmen.

35 13. Verfahren nach Anspruch 1, worin die zusätzlichen Monomeren, die zum Ersetzen der polymerisierten und als Produkt abgezogenen Monomeren zugefügt werden, in flüssiger Form unter dem Fließbett und bei oder nahe dem Eintrittspunkt der Zwei-Phasen-Mischung in den Reaktor eingeführt werden.

40 14. Verfahren nach Anspruch 1, worin die zusätzlichen Monomeren, die zum Ersetzen der polymerisierten und als Produkt abgezogenen Monomeren zugefügt werden, in den Rückflußstrom vor dem Eintritt der Zwei-Phasen-Mischung in den Reaktor eingeführt werden.

15. Verfahren nach Anspruch 1, worin Gasverteilungselemente unmittelbar unterhalb des Fließbettes und oberhalb des Bereichs vorgesehen sind, in welcher die Zwei-Phasen-Mischung in den Reaktor eingeführt wird.

45 16. Verfahren nach Anspruch 1, worin der Innendruck innerhalb des Reaktors etwa 0,69 MPa (100 psi) bis etw 2,4 MPa (350 psi) beträgt.

17. Verfahren nach Anspruch 1, worin das Monomer ein  $\alpha$ -Olefinmonomer mit 2 bis 4 Kohlenstoffatomen allein oder in Mischung mit einem oder mehreren anderen  $\alpha$ -Olefinmonomeren mit 2 bis 8 Kohlenstoffatomen ist.

## 50 Revendications

1. Procédé continu de production de polymère dans un réacteur à lit fluidisé à partir d'un ou plusieurs monomères fluides par passage continu d'un courant gazeux dans ledit réacteur en présence d'un catalyseur dans des conditions réactives, soutirage du produit polymérique et de fluides n'ayant pas réagi, refroidissement des fluides n'ayant pas réagi et retour des fluides refroidis dans le réacteur conjointement avec un complément suffisant de monomères pour remplacer les monomères polymérisés et soutirés comme produit, qui consiste à refroidir au-dessous du point de rosée une partie ou la totalité desdits fluides n'ayant pas réagi pour former un mélange de deux phases formé de gaz et de liquide entraîné, et à réintroduire ledit mélange de deux phases dans le réacteur.

2. Procédé suivant la revendication 1, dans lequel le mélange de deux phases formé de gaz et de liquide entraîné est introduit dans le réacteur en un point situé au-dessous du lit fluidisé.

3. Procédé suivant la revendication 1, dans lequel les fluides n'ayant pas réagi sont comprimés avant le refroidissement.

65 4. Procédé suivant la revendication 1, dans lequel des moyens sont prévus pour maintenir la phase

liquide du mélange de deux phases entraîné dans la phase gazeuse jusqu'au moment où ces liquides sont volatilisés ou entrent dans le lit fluidisé du réacteur.

5 5. Procédé suivant la revendication 1, dans lequel la vitesse du courant de gaz dans le réacteur est à un niveau ou au-dessus d'un niveau suffisant pour maintenir le lit dans un état de suspension et de fluidisation.

6. Procédé suivant la revendication 1, dans lequel un fluide condensable inert est présent dans le mélange de deux phases pour élever le point de rosée dudit mélange.

7. Procédé suivant la revendication 6, dans lequel le fluide inerte condensable est un hydrocarbure saturé.

10 8. Procédé suivant la revendication 1, dans lequel la phase liquide du mélange de deux phases représente moins d'environ 20% en poids du poids total dudit mélange de deux phases.

9. Procédé suivant la revendication 1, dans lequel la phase liquide dudit mélange de deux phases représente moins d'environ 10% en poids du poids total dudit mélange de deux phases.

15 10. Procédé suivant la revendication 1, dans lequel la vitesse d'écoulement du mélange de deux phases et le rapport du liquide au gaz dans le mélange sont maintenus à un niveau suffisant pour que le liquide entraîné reste dans ledit gaz jusqu'à ce qu'il soit volatilisé ou jusqu'à ce que ledit liquide entre dans le lit fluidisé du réacteur.

11. Procédé suivant la revendication 1, dans lequel le mélange de deux phases est fractionné en au moins deux courants séparés dont l'un au moins un introduit dans le réacteur en un point situé au-dessous du lit fluidisé et un ou plusieurs desdits courants sont facultativement introduits dans le lit fluidisé, pourvu que la vitesse de tous les gaz introduits au-dessous du lit fluidisé soit au moins suffisante pour supporter le lit et pour le maintenir dans un état fluidisé.

12. Procédé suivant la revendication 1, dans lequel le courant gazeux qui monte à travers le lit fluidisé est essentiellement uniforme quant à sa composition et traverse le lit de façon telle que le lit entier soit maintenu dans un état fluidisé et qu'il n'existe essentiellement aucun espace mort que des gaz ne puissent franchir.

13. Procédé suivant la revendication 1, dans lequel les monomères complémentaires ajoutés pour remplacer des monomères polymérisés et soutirés comme produit sont introduits dans le réacteur sous la forme liquide, au-dessous du lit fluidisé et au niveau ou à proximité du point d'entrée du mélange de deux phases dans le réacteur.

14. Procédé suivant la revendication 1, dans lequel les monomères complémentaires ajoutés pour remplacer les monomères polymérisés et soutirés comme produit sont introduits dans le courant recyclé avant l'entrée dudit mélange de deux phases dans le réacteur.

15. Procédé suivant la revendication 1, dans lequel un distributeur de gaz est prévu juste au-dessous du lit fluidisé et au-dessus de la région dans laquelle ledit mélange de deux phases est introduit dans le réacteur.

16. Procédé suivant la revendication 1, dans lequel la pression interne dans le réacteur va d'environ 0,69 MPa (100 lb/in<sup>2</sup>) à environ 2,4 MPa (350 lb/in<sup>2</sup>).

17. Procédé suivant la revendication 1, dans lequel le monomère est un monomère alpha-oléfinique ayant 2 à 4 atomes de carbone, seul ou en mélange avec un ou plusieurs autres monomères alpha-oléfiniques ayant 2 à 8 atomes de carbone.

45

50

55

60

65

